GPSA Engineering Data Book 14th Edition

REVISION	DATE	REASON(S) FOR REVISION
0	4/1/2017	Initial release

FIG. 19-1
Nomenclature

a'_t	=	tube flow area, ft^2	TS	=
a_t	=	total tube flow area, ft^2	U_D	=
A	=	absorption factor used in Equation 19-28	v	=
A_c	=	cross sectional area, ft^2	v_{max}	=
A_t	=	heat transfer surface, ft^2	v_i	=
AAM	=	tray active area, ft^2	v_o	=
ADM	=	tray downcomer area, ft^2	V	=
ATM	=	tower cross sectional area, ft^2	V_1	=
b	=	exponent used in Equations 19-5 and 19-6	V_{max}	=
B	=	bottoms product flowrate, moles/unit time	VD^*_{dsg}	=
C	=	coefficient in Equation 19.11, ft/hr	VD_{dsg}	=
CAF	=	vapor capacity factor, corrected, ft/sec	V_{load}	=
CAF_0	=	vapor capacity factor, uncorrected, ft/sec	V_0	=
CFS	=	vapor loading, ft^3/sec	w	=
D'	=	diameter, ft	x	=
D	=	distillate (overhead) product flowrate, moles/unit time	X	=
D_T	=	tower diameter, ft	X_{m+1}	=
E_a	=	absorption efficiency, Equation 19-30	x_1	=
E_s	=	stripping efficiency, Equation 19-32	X_0	=
f	=	friction factor (Moody friction factor/144), ft^2/in^2	y	=
F	=	feed rate, moles/unit time	Y_i	=
F_p	=	packing factor	Y_{n+1}	=
FF	=	flooding factor used in Equation 19-17, usually 0.82	Y_0	=
FPL	=	flow path length, ft	Z	=
g_c	=	conversion factor, 32.174 $(ft \cdot lb_m) / (lb_f \cdot sec^2)$		
G_t	=	mass velocity, $lb/(hr \cdot ft^2)$	**Greek**	
G_p	=	tower vapor loading, $lb/(ft^2 \cdot sec)$	α	=
GPM	=	tower liquid loading, gal/min	β_{ij}	=

H	=	enthalpy, Btu/lb
HETP	=	height of packing equivalent to a theoretical plate
HTU	=	height of a transfer unit
K	=	equilibrium K-value, y/x
L_0	=	liquid reflux rate, moles/unit time
L_p	=	liquid loading, lb/(ft^2·sec)
L	=	liquid rate, moles/unit time
L_t	=	tube length, ft
L_{m+1}	=	rich oil entering stripper, moles/unit time
LMTD	=	log mean temperature difference
m	=	number of stripping stages
M	=	mass flowrate, lb/hr
n	=	number of absorber stages
N_m	=	minimum number of theoretical stages
NP	=	number of passes in a tray
N_t	=	number of tubes
ΔP	=	pressure drop, psi
q	=	moles of saturated liquid in the feed per mole of feed
Q	=	heat transfer duty, Btu/hr
Q_c	=	condenser duty, Btu/hr
R	=	reflux ratio, moles of reflux divided by moles of net overhead product
Re	=	Reynold's number, dimensionless
s	=	specific gravity
S	=	number of stages
S_T	=	stripping factor used in Equation 19-31
S_F	=	separation factor defined by Equation 19-1
S_M	=	minimum number of stages defined by Equation 19-6

θ	=
σ	=
ρ	=
ε	=
μ	=

Subscripts

avg	=
B	=
BP	=
bottom	=
calc	=
corr	=
D	=
F	=
G	=
HK	=
i	=
L	=
LK	=
m	=
n	=
top	=
VF	=
v	=

tray spacing, inches

overall heat transfer coefficient, Btu/(hr·ft^2·°F)

specific volume, ft^3/lb

maximum velocity, ft/hr

specific volume of the inlet, ft^3/lb

specific volume of the outlet, ft^3/lb

vapor rate, moles/unit time

vapor rate leaving top tray, moles/unit time

volumetric vapor flow rate, ft^3/hr

downcomer velocity, uncorrected, gpm/ft^2

downcomer velocity, corrected, gpm/ft^2

vapor loading defined by Equation 19-13, ft^3/sec

stripping medium rate, moles/unit time

weight flow, lb/hr

liquid mole fraction

liquid rate, moles/unit time

moles of a component in the rich oil entering a stripper per mole of rich oil entering the stripper

moles of a component in the lean oil per mole of rich oil

moles of a component in the liquid in equilibrium with the stripping medium per mole of entering rich oil

vapor mole fraction

moles of any component in the lean gas leaving the absorber per mole of rich gas

moles of any component in the entering rich gas per mole of rich gas

moles of any component in the gas in equilibrium with the entering lean oil, per mole of rich gas

static head, ft

relative volatility

volatility factor defined in Equation 19-5

correlating parameter in Equations 19-7, 19-8
surface tension, dyne/cm
density, lb/ft^3
efficiency
viscosity, cp

average value
bottoms
bubble point feed stream
bottom of the column
calculated value
corrected value
distillate (overhead)
feed
gas
heavy key
any component
liquid
light key
minimum
tray number
top of the column
vaporized feed stream
vapor phase

Example 19-1

For the given feed stream, estimate the product stream compositions for 98% propane recovery in the overhead and with a maximum iso-butane content of 1%.

Feed	Component	Moles
	C_2	2.4
	C_3	162.8
	iC_4	31.0
	nC_4	76.7
	C_5	76.5
		349.4 moles

To find the propane in the overhead with 98% recovery:

C_3 in overhead = .98 * 162.8

159.5 moles

By steady state material balance, the moles of propane in the bottoms:

C_3 in bottoms = C_3 feed - C_3 overhead

3.3 moles

Because propane is the light key, all the ethane in the feed will appear in the overhead.

C_2 in overhead = C_2 in feed

2.4 moles

To find the amount of iso-butane in the overhead with maximum 1%:

The remainder of the materials in the overhead will be 99% of the total.

C_2	2.4 moles
C_3	159.5 moles
$C_2 + C_3$	161.9 moles

Total moles in overhead = 161.9 moles / .99

163.6 moles

Therefore, the number of moles of iso-butane can be found by:

iC4 in overhead= .01 * total moles in overhead

1.6 moles

The rest of the iso-butane will be in the bottoms.

iC_4 in bottoms = iC_4, feed - iC_4, overhead

29.4 moles

The remainder of the components (nC_4, C_5) will all be in the bottoms.
A material balance table is shown below.

	Feed		Overhead		Bottoms	
	moles	mole %	moles	mole %	moles	mole %
C_2	2.4	0.7	2.4	1.5	0.0	0.0
C_3	162.8	46.6	159.5	97.5	3.3	1.8
iC_4	31.0	8.9	1.6	1.0	29.4	15.8
nC_4	76.7	22.0	0	0.0	76.7	41.3
C_5	76.5	21.9	0	0.0	76.5	41.2
Total	349.4	100.0	163.6	100.0	185.8	100.0

Application of 19-1
This will calculate a material balance for the following components, given the information below.
User-entered data is in BOLD RED.

Operating Conditions and Design

Propane recovery in overhead		**0.98**
Maximum iso-butane content		**0.01**
Feed	**Component**	**Moles**
	C_2	**2.4**
	C_3	**162.8**
	iC_4	**31.0**
	nC_4	**76.7**
	C_5	**76.5**
		349.4 moles

Material Balance

C_3 in overhead	=	159.5	moles
C_3 in bottoms	=	3.3	moles
C_2 in overhead	=	2.4	moles
iC_4 in overhead	=	1.6	moles
iC_4 in bottoms	=	29.4	moles
nC_4 in bottoms	=	76.7	moles
C_5 in bottoms	=	76.5	moles

	Feed		Overhead		Bottoms	
	moles	mole %	moles	mole %	moles	mole %
C_2	2.4	0.7	2.4	1.5	0.0	0.0
C_3	162.8	46.6	159.5	97.5	3.3	1.8
iC_4	31.0	8.9	1.6	1.0	29.4	15.8
nC_4	76.7	22.0	0	0.0	76.7	41.3
C_5	76.5	21.9	0	0.0	76.5	41.2
Total	349.4	100.0	163.6	100.0	185.8	100.0

	$\alpha_{C3/iC4}$	=	1.643		

Tray Count

	α_{AVG}	=	1.855		Equation 19-4
	S_F	=	936.2		Equation 19-1
	S_m	=	11.08	trays	Equation 19-3
checking for changes in α	b	=	0.798		Equation 19-6
	β	=	1.759		Equation 19-6
	S_m	=	11.06	trays (with relative volatility correction)	
	S_m	=	11.08	trays (without relative volatility correction)	

Example 19-3

Find the diameter of a depropanizer given the following:

Vapor rate	70,418 ft^3/h	19.561 ft^3/s
Vapor density	3.0 lb/ft^3	
Liquid rate	1,190 gpm	
Liquid density	28.8 lb/ft^3	
Liquid surface tension	3.3 dyne/cm	
tray spacing	24 inches	

There are three methods of finding tower diameter. All three will be explored.

C Factor Method

From Figure 19.14, C was found to be approximately 425 ft/h.

Using Equation 19.11, v_{max} can be found.

v_{max} 1,261 ft/hr

This can be used in Equation 19.12 to find D.

D_T 8.43 feet
101 inches

Nomograph Method

V_{load} needs to be found in order to use Figure 19.15. V_{load} is found from Equation 19.13.

V_{load} 6.67 ft^3/s

Using Vload and the liquid rate of 1190 gpm on Figure 19.15, tower diameters were read for one and two pass tra

One pass tray	9.5 ft
	114 in
Two pass tray	7.5 ft

90 in

Detailed Method

From the equation in the bottom of Figure 19.16, the system factor for the tower was found.

System factor	0.85

Using Figure 19.17 and the given specifications, VD*dsg was found.

VD^*_{dsg}	186 gpm/ft^2

VDdsg = VD*dsg * System factor

VD_{dsg}	158.4 gpm/ft^2

From Figure 19.18, CAF_0 is 0.41.

$CAF = CAF_0$ * System factor

CAF	0.349 ft/s

Using D_t from the nomograph method for a one pass tray (9.25 ft) and a two pass tray (7.1 ft)

$FPL = 9 * D_T / NP$

FPL	85.5 ft	one pass tray
	33.75 ft	two pass tray

Using Equation 19.17, the active area can be found.

AAM	50.65 ft^2	one pass tray
	34.10 ft^2	two pass tray

The area of the downcomer can be found using Equation 19.18. If it is less than 11% of AAM, use either 11% of
double ADM, whichever is smaller.

ADM	9.16 ft^2
ADM / AAM	0.18 one pass
	0.27 two pass

In both cases, the downcomer areas are sufficiently large.

Now the cross sectional area of the tower can be found using Equation 19.19

ATM	68.98 ft^2	one pass tray
	52.43 ft^2	two pass tray

Another method to find the cross sectional area of the tower is Equation 19.20.

ATM	29.88 ft^2

The larger of the two ATM values is used. In this case, it will be the ones calculated from Equation 19.19.
The diameter of the column can be calculated by Equation 19.21.

D_T	9.37 ft	one pass tray
	8.17 ft	two pass tray

A comparison of the different calculated diameters follows.

Method	**Number of Passes**	**Estimated Diameter (in)**
C Factor	-	101
Nomograph	1	114
Nomograph	2	90
Detailed	1	112
Detailed	2	98

These calculation spreadsheets are provided to provide an “Operational level” of accuracy calculation based on r ays.

Application of 19-3

This finds the diameter of a depropanizer with the following speci

User-entered data is in BOLD RED.

Operating Conditions and Design

Vapor rate	=	**70,418**	ft^3/h
	=	19.561	ft^3/s
Vapor density	=	**3.0**	lb/ft^3
Liquid rate	=	**1,190**	gpm
Liquid density	=	**28.8**	lb/ft^3
Liquid surface tension	=	**3.3**	dyne/cm
tray spacing	=	**24**	inches

C Factor Method

Using the surface tension and tray spacing entered above, use I

C	=	**430**	ft/h
v_{max}	=	1,261	ft/h
D_T	=	8.43	feet

Nomograph Method

V_{load}	=	6.67	ft^3/s
D_T - One Pass Tray	=	**9.5**	ft
Two Pass Tray	=	**7.5**	ft

Detailed Method

System factor	=	0.85	
ρ_V-ρ_L	=	25.8	lb/ft^3
VD^*_{dsg}	=	**186**	gpm/ft^2
VD_{dsg}	=	158.4	gpm/ft^2
CAF_0	=	**0.41**	ft/s
CAF	=	0.349	ft/s
FPL - One Pass	=	85.5	ft
Two Pass	=	33.75	ft

FF	=	**0.82**	
AAM - One Pass	=	50.65	ft^2
Two Pass	=	34.10	ft^2

The area of the downcomer can be found using Equation 19.18. If
either 11% of AAM or double ADM, whichever is smaller.

ADM	=	9.16	ft^2
ADM / AAM: -One Pass	=	0.18	
Two Pass	=	0.27	
ATM - One Pass	=	68.98	ft^2
Two Pass	=	52.43	ft^2
ATM	=	29.88	ft^2

The larger of the two ATM values is used. In this case, it will be th

D_T - One Pass	=	9.37	ft
Two Pass	=	8.17	ft

Method	Number of Passes	Estimated Diameter (in)
C Factor	-	101
Nomograph	1	114
Nomograph	2	90
Detailed	1	112
Detailed	2	98

AAM or

in the Engineering Data Book as published by the Gas Processors Suppliers Association as a service
eets based on the GPSA Engineering Data Book sample calculations, the use of such information is v
sonableness of factual or scientific assumptions, studies or conclusions, or merchantability, fitness for
mitation, those resulting from lost profits, lost data or business interruption) arising from the use, ina

ather broad assumptions (including but not limited to: temperatures, pressures, compositions, imperia fications using three methods.

Figure 19.14 and input C.

Figure 19-14

Equation 19-11

Equation 19-12

Equation 19-13

Figure 19-15

Figure 19-15

Figure 19-16

Figure 19-17

Equation 19-14

Figure 19-18

Equation 19-15

Equation 19-16

Equation 19-17
Equation 19-17

it is less than 11% of AAM, use

must be at least 11% of AAM

Equation 19-19

Equation 19-20

e ones calculated from Equation 19-19.

Equation 19-21
Equation 19-21

ıl curves, site conditions etc) and do not replace detailed and accurate Design Engineering taking into accou

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:rade-name, trademark, and service mark manufacturer or otherwise does not constitute or imply endorsement, rec and limitations.

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Example 19-4

Find the tray efficiency of the column in Example 19-2.

Figure 19.18 will be used to estimate a plate efficiency. This needs the relative volatility and the viscosity of the key component at average column conditions.

$$T_{avg} = (T_{top} + T_{bottom}) / 2$$

Tavg 185 °F

It is given that at 185 F, the viscosity of the feed is 0.076 cp and the average α is 1.854.

μ	0.076	cp
α	1.854	

To use Figure 19.19, the product of these two is needed.

product 0.141

From the figure, the efficiency was estimated to be 80%. The number of actual trays can be found with this number as follows. The method below counts the reboiler as a stage.

$$N_{trays} = (\text{theoretical trays - 1}) / \text{efficiency}$$

N_{trays} 25

These calculation spreadsheets are provided to provide an “Operational level” of accuracy calculation based on ra

Application of 19-4

This will find the tray efficiency of a column.

User-entered data is in BOLD RED

Operating Conditions and Design

T_{top}	=	**120**	°F
T_{bottom}	=	**250**	°F
T_{avg}	=	185	°F

Tray Efficiency

μ	=	**0.076**	cp	
α	=	**1.854**		
μ*α	=	0.141		
ε	=	**0.8**		Figure 19-19
$N_{theoretical\ trays}$	=	**21**	trays	
N_{trays}	=	**25**	trays	Equation 19-22

.ther broad assumptions (including but not limited to: temperatures, pressures, compositions, imperial curves, site

conditions etc) and do not replace detailed and accurate Design Engineering taking into account actual process co

nditions, fluid properties, equipment condition or fowling and actual control set-point dead-band limitations.

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Example 19-5

Find the diameter for a packed tower using 2" plastic Pall rings for the column in Example 19-3. The g
for that problem are copied below.

Vapor rate	70,418 ft^3/h	19.561 ft^3/s
Vapor density	3 lb/ft^3	
Liquid rate	1,190 gpm	
Liquid density	28.8 lb/ft^3	
Liquid surface tension	3.3 dyne/cm	
tray spacing	24 inches	

Also given:

μ	0.076 cp
ΔP	0.5 in H_2O/ft packing

From Figure 19.26, the packing factor (F_p) for the specified packing is 26.

F_p 26

Figure 19.27 can now be used. The bottom axis is defined by (L_p / G_p) * sqrt (ρ_v / ρ_L). L_p/ G_p can be s

M_L =(1190 gpm * 18.8 lb/ft^3 * 60 min/h)/ (7.48 gal/ft^3)

M_L 274,909 lb/h

M_G = 70,418 ft^3/h * 3 lb/ft^3

M_G 211,254 lb/h

bottom axis of Figure 19.27 0.420

Using 0.420 on the bottom axis, following the graph up to the specified pressure drop, the left axis can

Using 0.420 on the bottom axis, following the graph up to the specified pressure drop, the left axis can
The left axis is equal to a large equation that includes Gp, which can be solved for.

left axis	0.024
Gp	1.659 lb/ft^2*s

The cross sectional area of the column can be found by taking the mass of the gas flowrate and dividing by Gp and the conversion between seconds and hours.

Ac	35.37 ft^2

The diameter of the tower can be found using the equation for area of a circle.

D_T	6.71 ft

This would likely be rounded to 7 ft

iven data

substituted with M_L / M_G.

be found

Application of 19-5

This will find the diameter of a packed tower.

User-entered data is in BOLD RED.

Operating Conditions and Design

Vapor flow rate	=
	=
Vapor density	=
Liquid rate	=
Liquid density	=
Liquid surface tension	=
tray spacing	=
μ	=
ΔP	=
Packing Type	=

Diameter

F_p	=
$Mass_L$	=
$Mass_G$	=
Horizontal Axis of Figure 19-27	=
Vertical Axis of Figure 19-27	=
G_p	=
Ac	=
D_T	=

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spreadsheets based on the GPSA Engineering Data Book sample calculations, the use of such information is
acy or reasonableness of factual or scientific assumptions, studies or conclusions, or merchantability, fitness f
without limitation, those resulting from lost profits, lost data or business interruption) arising from the use, i
ased on rather broad assumptions (including but not limited to: temperatures, pressures, compositions, imper

70,418	ft^3/h	
19.561	ft^3/s	
3	lb/ft^3	
1,190	gpm	
28.8	lb/ft^3	
3.3	dyne/cm	
24	inches	
0.076	cp	
0.5	in H_2O/ft packing	
	2" plastic Pall Rings	
26		Figure 19.26
274,909	lb/h	
211,254	lb/h	
0.420		
0.024		Figure 19-27
1.659	lb/ft^2*s	Figure 19-27
35.368592	ft^2	
7	ft	

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Example 19-6

Find the optimum heat exchanger for a vertical thermosyphon application given it must produce 40,800 lb/h va
pressure of the column is 275 psig, has an isothermal boiling point of 228 F. The energy for the reboiler will be
125 psig. The recirculation ratio should be at least 4:1.

Tube Data

Inner Diameter	0.62	in
Surface Area	0.1963	ft^2/ft
Internal Tube Area	0.302	in^2
Vapor Density	2.27	lb/ft^3
Liquid Viscosity	0.1	cp
Liquid Specific Gravity	0.43	

An energy balance should be calculated. Using thermodynamic data:

Enthalpy at 228 F and 290 psia of:

Liquid Butane	241	Btu/lb
Vapor Butane	338	Btu/lb

The energy needed to make 40,800 lb/ vapor:

$Q = m * \Delta H$

Q	3,957,600	Btu/h

To find how much steam is needed:

Enthalpy of steam at given conditions	868	Btu/lb

m= Q/ H

m	4,559	lb/h steam
ΔT	125	°F

The maximum energy flux is 12,000 Btu/ft^2

$A = Q / flux$

A	329.8 ft^2

This is the area that the heat exchanger must provide. Using the specification for the surface area and assuming the number of tubes can be found. I'll start using 16 ft, 12 ft, and 10 ft long tubes.

	Length of Tube		
	16 ft	12 ft	10 ft
N = required surface area / (length of tube * surface area of tube)			
N	105 tubes	140 tubes	168 tubes
Using Equation 19.27, the static pressure of the reboiler leg can be found.			
$v_v = 1 / \rho_v$			
v_v	0.4405 ft^3/lb	0.4405 ft^3/lb	0.4405 ft^3/lb
$v_L = 1 / \rho_L$			
v_L	0.0373 ft^3/lb	0.0373 ft^3/lb	0.0373 ft^3/lb
The weight of the recirculated liquid can be found by multiplying the vapor mass by 4 (the minimum recircula			
M_L	163,200 lb liquid/hr	163,200 lb liquid/hr	163,200 lb liquid/hr
With this, the total volume of the reboiler outlet can be found.			
$V_L = M_L * v_L$			
V_L	6082 ft^3	6082 ft^3	6082 ft^3
$V_v = M_v * v_v$			
Vv	17,974 ft^3	17,974 ft^3	17,974 ft^3
Total volume	24,056 ft^3	24,056 ft^3	24,056 ft^3

The specific volume of the outlet can now be found using the total volume and the total mass.

$v_o = V_o / M_o$

v_o	0.1179 ft^3/lb	0.1179 ft^3/lb	0.1179 ft^3/lb

Now that v_o has been obtained, the static pressure of the reboiler leg can be found using Eq. 19.27.

P	1.59 psi	1.19 psi	0.99 psi

Now the frictional resistance can be found. When added to the static pressure, this will give the total resistance
frictional resistance to flow, the area of flow must be found. This is $a_t = N_t * a'_t / 144$

a_t	0.220 ft^2	0.294 ft^2	0.352 ft^2

The mass velocity, Gt, can be found by dividing the mass flowrate by the area of the tube.

G_t	926,351 lb/hr * ft^2	694,763 lb/hr * ft^2	578,969 lb/hr * ft^2

Converting the viscosity units:

μ	0.242 ft*h/lb	0.242 ft*h/lb	0.242 ft*h/lb

Converting the diameter of the pipe:

D	0.052 ft	0.052 ft	0.052 ft

Now the Reynold's number can be found. $Re = D * G_t / \mu$

Re	197,775	148,331	123,609

Using a Moody plot, the friction factor can be found.

f	0.000127 ft^2/in^2	0.000135 ft^2/in^2	0.0001483

The average specific gravity can be found.

s_{avg}	0.283	0.283	0.283

Using Bernoulli's equation the pressure drop can be found.

ΔP	2.28 psi	1.02 psi	0.65 psi

The total resistance to flow can be calculated by adding the frictional resistance and static resistance.

Total ΔP	3.87 psi	2.21 psi	1.64 psi

The driving force can be calculated:

	2.98 psi	2.24 psi	1.86 psi

The difference in the driving force and resistance to flow determines whether or not the flow will go into the r

	-0.89 psi	0.02 psi	0.22 psi

apor (assume pure butane). The
e supplied by saturated steam at

Application of 19-6

This will find the optimum heat exchanger tu

User-entered data is in BOLD RED.

Operating Conditions and Design

Saturated steam heat supply =

Tube Data

Inner Diameter	=
Surface Area	=
Internal Tube Area	=
Vapor Density	=
Liquid Viscosity	=
Liquid Specific Gravity	=

Thermodynamic Data and Energy Balanc

Isothermal Boiling Point:	=
Column Pressure:	=
Enthalpy of bottoms liquid at specified T, P	=
Enthalpy of bottoms vapor at specified T, P	=
Enthalpy of steam at given conditions	=
Maximum Energy Flux	=
Q	=
m	=
ΔT	=
A	=

Tube Calculations

g a length,

tion ratio).

Length of Tube =

N =

$v_v = 1/\rho_v$ =

$1/\rho_L$ =

M_L =

V_L =

Vv =

Total volume =

v_o =

P =

a_t =

G_t =

μ =

D' =

Re =

f =

s_{avg} =

ΔP_t =

Total ΔP =

Driving Force =

difference =

GRAPHIC SOLUTION

To find the length where the difference is zer
where the curve intersects with the y axis to
outside the range.

Length of Tubes, ft	Difference
10	
12	
14	
16	
18	
20	
22	

GOAL SEEKING SOLUTION

To find the length where the difference is zer
the What-If Analysis button and selecting Go
click OK. Cell P34 will show the maximum l

e to flow. To find the

Length of Tube, ft	Difference
12.51	0

eboiler.

ed in the Engineering Data Book as published by the Gas Processors Suppliers Association as a servic
lsheets based on the GPSA Engineering Data Book sample calculations, the use of such information is
easonableness of factual or scientific assumptions, studies or conclusions, or merchantability, fitness f
t limitation, those resulting from lost profits, lost data or business interruption) arising from the use, ir
 rather broad assumptions (including but not limited to: temperatures, pressures, compositions, imper

ıbe length for a vertical thermosyphon.

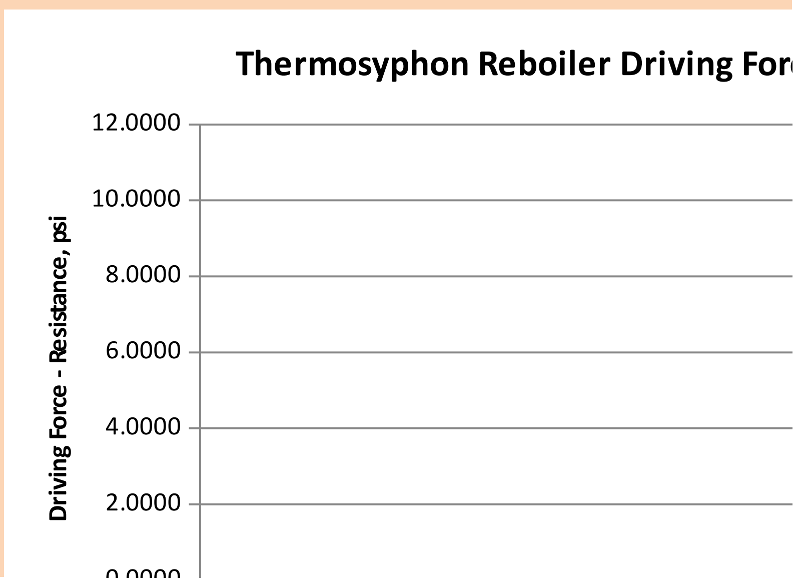
40,800	lb vapor produced/h		
4	:1 recirculation ratio		recirculation ratio must be greater than or equal t
125	psig	**353**	°F
0.62	in		
0.1963	ft^2/ft		
0.302	in^2		
2.27	lb/ft^3		
0.1	cp		
0.43			

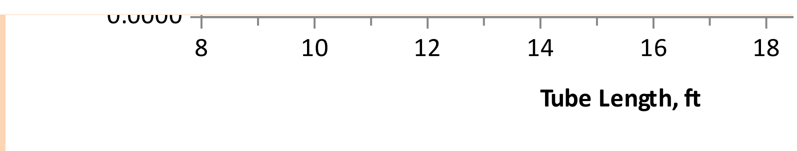
e

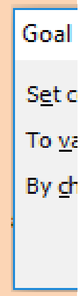
228	°F	Isothermal boiling point at column pressure
290	psia	
241	Btu/lb	
338	Btu/lb	
868	Btu/lb	
12,000	Btu/ft^2	
3,957,600	Btu/h	
4559	lb/h steam	
125	°F	Log Mean Temperature Difference
329.8	ft^2	

16	ft	Starting assumption
105	tubes	
0.4405	ft³/lb	
0.0373	ft³/lb	
163200	lb liquid/hr	
6,082	ft³	
17,974	ft³	
24,056	ft³	
0.1179	ft³/lb	
1.59	psi	Equation 19-27
0.220	ft²	
926,351	lb/hr * ft²	
0.242	ft*h/lb	
0.052	ft	
197,775		
0.000127	ft²/in²	Moody plot
0.283		
2.28	psi	
3.87	psi	
2.98	psi	
-0.89	psi	

o, use this curve. Enter a length in cell N64. The table will populate with lengths plus or minus 2 feet and
find the maximum length where the required recirculation ratio is assured. If there is no intersect, choose





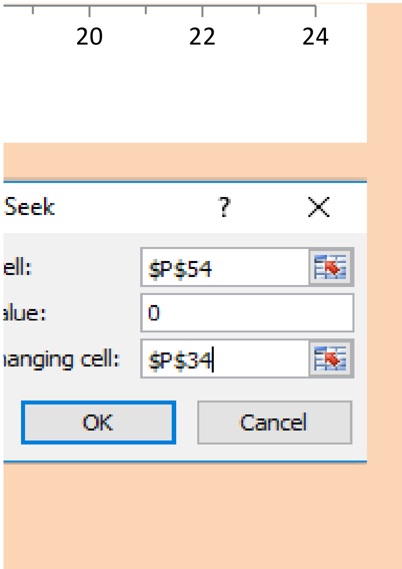


ro, use Goal Seek. This is found on the Data Ribbon, in the Data Tools group by pulling down
oal Seek. Fill in the pop-up form with the information shown in the figure to the right and
length where the required recirculation ratio is assured.

o 4:1

calculate the difference. Find
a difference value for N64

ce Curve


20
22
24
Seek
?
ell:
P54
alue:
0
anging cell:
P34
OK
Cancel

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such information. Reference herein to any specific commercial product, calculation method, process, or service b

ract, tort or any other legal theory and whether or not advised of the possibility of such damages.
count actual process conditions, fluid properties, equipment condition or fowling and actual control set-point dead

y trade-name, trademark, and service mark manufacturer or otherwise does not constitute or imply endorsement, r

l-band limitations.

:ecommendation or favoring by the GPA and/(

GPSA Engineering Data Book 14th Edition

Example 19.17

Find the oil circulation rate and the composition of the residue gas given the following information. 75 p
from 100 mol of the rich gas stream. The absorber will have six theoretical plates, the average temperatu
and 1000 psig. Assume the lean oil is completely stripped of rich gas components. The feed composition

Component	Mol %
C_1	90.6
C_2	4.3
C_3	3.2
iC_4	0.5
nC_4	1.0
C_6	0.4

K values can be found from the equilibrium data in Chapter 25, using the average absorber conditions.

Component	K
C_1	3.250
C_2	0.900
C_3	0.370
iC_4	0.100
nC_4	0.170
C_6	0.035

From Figure 19.52, A can be found using Ea as .75 (specified efficiency for propane absorption) and n=6

A	0.8

Equation 19.29 can now be used.

L_0	29.6	mol/h

Now, A can be calculated for the remaining components.

Component	A
C_1	0.091
C_2	0.329
C_3	0.800

iC_4	2.960
nC_4	1.741
C_6	8.457

Now the absorption efficiencies can be determined for each component, using Figure 19.52.

Component	Ea
C_1	0.091
C_2	0.329
C_3	0.75
iC_4	0.96
nC_4	0.98
C_6	1

Now, the Ea value can be used to solve Eq 19.30 for the outlet composition of the lean gas.

Component	Y_1
C_1	82.36
C_2	2.89
C_3	0.80
iC_4	0.02
nC_4	0.02
C_6	0.00

Now the moles of each component in the rich oil, I, can be calculated by steady state material balance.

Component	I
C_1	8.24
C_2	1.41
C_3	2.40
iC_4	0.48
nC_4	0.98
C_6	0.40

All the calculated properties are summarized in a table below.

Component	**Mol %**	**K**	**A**	**Ea**	**Y1**	**I**
C_1	90.6	3.250	0.091	0.091	82.36	8.24
C_2	4.3	0.900	0.329	0.329	2.89	1.41
C_3	3.2	0.370	0.800	0.75	0.80	2.40

iC_4	0.5	0.100	2.960	0.96	0.02	0.48
nC_4	1.0	0.170	1.741	0.98	0.02	0.98
C_6	0.4	0.035	8.457	1	0.00	0.40
TOTAL	100.0				86.08	13.92

ercent of the propane needs to be removed
ure and pressure of the absorber are 104 F
is given below.

(specified trays).

Application of 19.17

This will find the oil recirculation ra
completely stripped.

User-entered data is in BOLD RE

Component	Mol %
C_1	**90.6**
C_2	**4.3**
C_3	**3.2**
iC_4	**0.5**
nC_4	**1.0**
C_6	**0.4**

K values can be found from the eq

Component	K
C_1	**3.250**
C_2	**0.900**
C_3	**0.370**
iC_4	**0.210**
nC_4	**0.170**
C_6	**0.035**

A = **0.8**
L_0 = 29.6

Component	A
C_1	0.091
C_2	0.329
C_3	0.800
iC_4	1.410
nC_4	1.741
C_6	8.457

Component	Ea

C_1	**0.091**
C_2	**0.329**
C_3	**0.75**
iC_4	**0.96**
nC_4	**0.985**
C_6	**1**

Component	Y_1
C_1	82.36
C_2	2.89
C_3	0.80
iC_4	0.02
nC_4	0.01
C_6	0.00

Component	l
C_1	8.24
C_2	1.41
C_3	2.40
iC_4	0.48
nC_4	0.99
C_6	0.40

All the calculated properties are sum

Component	**Mol %**
C_1	90.6
C_2	4.3
C_3	3.2
iC_4	0.5
nC_4	1.0
C_6	0.4
Total	100.0

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preadsheets based on the GPSA Engineering Data Book sample calculations, the use of such information
y or reasonableness of factual or scientific assumptions, studies or conclusions, or merchantability, fitnes
ithout limitation, those resulting from lost profits, lost data or business interruption) arising from the use
sed on rather broad assumptions (including but not limited to: temperatures, pressures, compositions, im

ate and the composition of the residue gas for an absorber, assuming the lean oil comes in

D.

75 % of propane removed

6 theoretical plates

104 F, average temperature

1000 psig, average pressure

quilibrium data in Chapter 25, using the average absorber conditions. Enter them below.

Figure 19-52

mol/h Equation 19-29

Equation 19-28

Figure 19-52

Equation 19-30

nmarized in a table below.

K	A	Ea	Y1	l
3.250	0.091	0.091	82.36	8.24
0.900	0.329	0.329	2.89	1.41
0.370	0.800	0.75	0.80	2.40
0.210	1.410	0.96	0.02	0.48
0.170	1.741	0.985	0.01	0.99
0.035	8.457	1	0.00	0.40
			86.1	13.9

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information. Reference herein to any specific commercial product, calculation method, process, or service by trac

ort or any other legal theory and whether or not advised of the possibility of such damages.
actual process conditions, fluid properties, equipment condition or fowling and actual control set-point dead-band

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l limitations.

mendation or favoring by the GPA and/or GP

Example 19-8

Determine the number of theoretical stages given the following information. Sour water containing 2500 pp ppmw. Enough energy is provided by the reboiler to produce .75 lb steam per gallon feed. The feed rate is 1 operates at 21 psia. First, an overall material balance will be performed using the given specifications.

The feed will be converted to mass flowrate.

Feed	10	gpm
	4,998	lb/h

Now the mass of overhead steam can be calculated using the given specification.

Overhead steam	450	lb/h

Now the overall steady state material balance can be done, using the specifications given.

	Feed (lb/h)	Overhead (lb/h)	Bottoms (lb/h)
H_2S	12.495	12.488	0.007
H_2O	4,985.505	450.000	4,535.505
total	4,998.000	462.488	4,535.512

The fraction of H_2S stripped can be found by dividing the H_2S in the overhead by the H_2S in the feed.

Fraction H_2S Stripped

0.9995

In order to estimate the top temperature, the fraction of water in the overhead and the partial pressure of water in the overhead need to be found.

Fraction of H_2O in overhead

0.973

Partial pressure of H_2O = fraction of H_2O * pressure

20.433 psi

Using the steam tables from Chapter 24, the temperature of the top was estimated to be 229 °F.

Now that the temperature is known, the K value for H_2S can be obtained. K = H / P where H is Henry's Lav

From Figure 19.53:

T (°F)	H, H_2S (psia)
100	11,000
200	18,200
300	26,000

At 229 °F, the Henry's constant was interpolated and found to be 2.05 10^4.

K 976.19

Now the moles of vapor leaving the top tray can be found using the masses from the material balance and th

V 25.37 mol

The same can be done for the moles of liquid leaving the top tray.

L 277.34 mol

Now the stripping factor can be found. ST = K * V / L

S_T 89.29

Now various values of Es, the efficiency, can be calculated assuming multiple values of m.

m	Es
1	0.98892
2	0.99988
3	1.00000

In order to get the required H_2S removal fraction, 2 theoretical trays are needed.

mw needs to be stripped to 1.5
0 gpm and the top of the tower

v Constant and P is the pressure.

ne molecular weights of each component.

shed in the Engineering Data Book as published by the Gas Processors Supplier
eadsheets based on the GPSA Engineering Data Book sample calculations, the us
r reasonableness of factual or scientific assumptions, studies or conclusions, or n
out limitation, those resulting from lost profits, lost data or business interruption
l on rather broad assumptions (including but not limited to: temperatures, pressu

Application of 19-8

This will calculate the number of theoretical stages needed to strip H_2S from sour water.

User-entered data is in BOLD RED.

Feed	=	**10**	gpm	
	=	4,998	lb/h	
H_2S Concentration, inlet	=	**2,500**	ppmw	
outlet	=	**1.5**	ppmw	
operating pressure	=	**21**	psia	typically between 24.7 to 29.7
steam produced per gallon feed	=	**0.75**	lb	

Mass Balance

Overhead steam = 450 lb/h

	Feed (lb/h)	Overhead (lb/h)	Bottoms (lb/h)
H_2S	12.495	12.488	0.007
H_2O	4,985.505	450.000	4,535.505
total	4,998.000	462.488	4,535.512

Fraction H_2S Stripped	=	0.9994	
Mass Fraction H_2O in Overhead	=	0.973	
Partial Pressure H_2O	=	20.433	psi

Stripping Calculation

Using the steam tables from Chapter 24, estimate the temperature of the top and enter it below.

T_{top}	=	**229**	°F	
H	=	**20,575**	psia	Figure 19-53
K	=	979.8		
V	=	25.37	mol	
L	=	277.34	mol	
S_T	=	89.615		Equation 19-31

m	E_s	Equation 19-32

1	0.98896	
2	0.99988	
3	1.00000	Required fraction
4	1.00000	0.9994
5	1.00000	

psia

, process, or service by trade-name, trademark, and service mark manufacturer or otherwise does not constitute or

ıges.

control set-point dead-band limitations.

imply endorsement, recommendation or favoring by the GPA and/or

LIMITS

Example 19-2
Optimum operating reflux ratio are 1.2 to 1.3 times the minimum reflux ratio

Example 19-3
Use flooding factor of 0.82 for most systems

Example 19-4
Add an extra tray to tray count for each feed tray and side exchanger

Example 19-5
Pressure drop for packed columns should be 0.20 to 0.60 inches of water per foot of pack depth; 1 inch maximum

Example 19-6
Use recirculation ratios 4:1 or greater
Use the maximum allowable flux when initially determining reboiler surface area

Example 19-7
Use average absorption factor determined by Kremser and Brown, eqn. 19-28,29

Example 19-8
Typical operating conditions:

Pressure	10-15 psig
Feed Temp.	200-230 °F
Bottoms Temp.	240-250 °F
Reboil Heat	1000-2000 Btu/gal
Residual H2S	0.5-2.0 ppmw

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